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TECHNOECONOMIC ANALYSIS OF A LOW CO_2 EMISSION DIMETHYL ETHER (DME) PLANT BASED ON GASIFICATION OF TORREFIED BIOMASS

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Abstract

Two models of a DME fuel production plant were designed and analyzed in DNA and Aspen Plus. The plants produce DME by either recycle (RC) or once through (OT) catalytic conversion of a syngas generated by gasification of torrefied woody biomass. Torrefication is a mild pyrolysis process that takes place at 200-300°C. Torrefied biomass has properties similar to coal, which enables the use of commercially available coal gasification processing equipment. The DME plants are designed with focus on lowering the total CO_2 emissions from the plants; this includes e.g. a recycle of a CO_2 rich stream to a CO_2 capture plant, which is used in the conditioning of the syngas.

The plant models predict energy efficiencies from torrefied biomass to DME of 66% (RC) and 48% (OT) (LHV). If the exported electricity is included, the efficiencies are 71% (RC) and 64% (OT). When accounting for energy loss in torrefaction, the total efficiencies are reduced to 64% (RC) and 58% (OT). The two plants produce DME at an estimated cost of $\$11.9/GJ_{LHV}$ (RC) and

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$12.9/GJ_{LHV}$ (OT). If a credit is given for storing the CO_2 captured, the future costs may become as low as $5.4/GJ_{LHV}$ (RC) and $3.1/GJ_{LHV}$ (OT).

Keywords: biorefinery, biofuel, dimethyl ether, DME, torrefication, gasification, syngas, CO_2 capture.

1. Introduction

One of the ways of reducing the CO_2 emissions from the transportation sector is by increasing the use of biofuels in vehicular applications. Dimethyl ether (DME) is a diesel-like fuel that can be produced from biomass in processes very similar to methanol production processes. Combustion of DME produces lower emissions of NO_x than combustion of diesel, with no particulate matter or SO_x in the exhaust [1], however it also requires storage pressures in excess of 5 bar to maintain a liquid state (this pressure is similar to LPG). Other "advanced" or "second generation" biofuels include methanol, Fischer–Tropsch diesel and gasoline, hydrogen and ethanol. Like DME and methanol, Fischer–Tropsch fuels and hydrogen are also produced by catalytic conversion of a syngas[1]. Ethanol could also be produced by catalytic conversion of a syngas (at research stage), but is typically produced by biological fermentation. Of these fuels, only hydrogen can be produced at a higher biomass to fuel energy efficiency than methanol and DME. Ethanol (produced from fermentation of cellulosic biomass) and Fischer–Tropsch fuels have lower biomass to fuel energy efficiency than methanol and DME [2]. The advantage of Fischer–Tropsch diesel and gasoline – as well as methanol and ethanol blended in gasoline - is that these fuels can be used

[1] For hydrogen, the catalytic conversion occurs in a water gas shift (WGS) reactor, where steam reacts with CO to produce hydrogen. Hydrogen can also be produced by fermentation.

in existing vehicle power trains, while hydrogen, DME and neat ethanol and methanol require new or modified vehicle power trains.

The relative low cost needed to implement DME as a transportation fuel, together with its potential for energy efficient production and low emissions (including low well-to-wheel greenhouse gas emissions) when used in an internal combustion engine, makes DME attractive as a diesel substitute [2].

Two DME production plants, based on syngas from gasification of torrefied wood pellets, are investigated in this paper:

- The OT plant uses once through synthesis and the unconverted syngas is used for electricity production in a combined cycle.
- The RC plant recycles unconverted syngas to the DME reactor to maximize DME production.

Both plants uses CO_2 capture to condition the syngas for DME synthesis and the captured CO_2 is sent to underground storage. The plants are designed with focus on lowering the total CO_2 emission from the plants, even though the feedstock used is biomass. Capturing and storing CO_2 from a biomass plant, gives a negative greenhouse effect, and can be an interesting concept, if a credit is given for storing CO_2 generated from biomass. The concept of receiving a credit for storing CO_2 generated from biomass has been investigated before (e.g., in [3]), but a study of the thermodynamics and economics of a biomass-based liquid fuels plant, where the focus in the design of the plant, was lowering the total CO_2 emission from the plant is not presented in the literature.

The DME plants modeled are of large-scale (> 2,000 tonnes per day) because of the better economics compared to small-scale production of DME [3,4]. Larger–scale plants, however, have higher feedstock transportation costs, which increase the attractiveness of torrefied wood pellets as

a feedstock instead of conventional wood pellets. Torrefication of biomass also makes it possible to use commercially available coal gasification processing equipment[2].

Production of DME from biomass has been investigated before (e.g., [5,6]). In [6] the feedstock used is black liquor and in [5] the feedstock used is switchgrass.

This paper documents the design of two DME plants using DNA[3] [7,8] and Aspen Plus modeling tools. Thermodynamic and economic performance of the plant configurations are presented and discussed.

1.1 Torrefication of biomass

Torrefaction of biomass is a mild pyrolysis process, taking place at 200-300°C. The process alters the properties of biomass in a number of ways, including increased energy density, improved grindability/pulverization, better pelletization behavior, and higher resistance to biodegradation and spontaneous heating. This conversion process enables torrefied biomass to achieve properties very similar to coal, and therefore allows the altered biomass feedstock to be handled and processed using conventional coal preparation methods. Additionally, torrefied biomass can be stored in outdoor environments and the electricity consumption for milling and pelletization is significantly lower than that of wood [9,10].

2. Design of the DME plant

A simplified process flow sheet of the DME plant design is shown in Fig. 1 and detailed process flow sheets are shown in Fig. 5 and Fig. 6. Plant design aspects related to feedstock preparation,

[2] See the Gasification World Database [11] for a list of commercial gasification plants.

[3] Because of DNA's excellent solids handling, DNA was used to model the gasifier. The rest of the modeling was done with Aspen Plus.

gasification, syngas conditioning, DME synthesis and distillation are described next and are followed by a discussion of electricity co-production in the two plants and the commercial status of the process components used. Important process design parameters used in the modeling are shown in Table 1.

Pretreatment & feeding

The pretreatment and feeding of torrefied wood pellets are assumed to be accomplished with existing commercial coal technology [9,10]. The torrefied biomass is milled to powder and the powder is pressurized with lock hoppers and fed to the gasifier with pneumatic feeders, both using CO_2 from the carbon capture process downstream.

Gasification

A commercial, dry-fed, slagging[4] entrained flow coal gasifier from Shell is used for gasifying the torrefied wood powder. The gasifier is oxygen blown, pressurized to 45 bar and steam moderated [12]. The oxygen supply is provided by a cryogenic air separation unit. A gas quench using about 200°C recycled syngas downstream of the dry solids removal lowers the temperature of the syngas from 1300°C to 900°C. The composition of the syngas is calculated by assuming chemical equilibrium at 1300°C (composition given in Table 2 and Table 3).

Gas cooling and water gas shift

The syngas is further cooled to 200-275°C by generating superheated steam for primarily the integrated steam cycle[5]. A sulfur tolerant[6] water gas shift (WGS) reactor adjusts the H_2/CO ratio

[4] Because of the low ash content in biomass a slag recycle is needed to make the gasifier slagging [13]. Also see note b below Table 1.

[5] Steam is superheated to 600°C in the gas cooling (at 55 bar (RC) or 180 bar (OT)). In [12] it is stated that only a "mild superheat" can be used in the gas cooling, but in [14] steam at 125 bar is superheated to 566°C.

to 1 (RC plant) or 1.6 (OT plant). In the RC plant, the H_2/CO ratio is adjusted to 1, to optimize DME synthesis according to Eq. 1 [5]. In the OT plant, the H_2/CO ratio is set to 1.6 to increase the amount of CO_2 captured in the downstream conditioning and thereby minimizing the CO_2 emissions from the plant. After the WGS reactor, the gas is cooled to 30°C prior to the acid gas removal step.

Gas cleaning incl. CCS

Gas cleaning of biomass syngas for DME synthesis includes cyclones and filters for particle removal placed just after the high temperature syngas cooler, an acid gas removal (AGR) step and guard beds[7] placed just before the synthesis reactor [15,16]. The AGR step is done with a chilled methanol process similar to the Rectisol process [17,18], and it removes sulfur components (H_2S and COS[8]), CO_2 and other species such as NH_3 and HCl in one absorber (Fig. 2). By using only one absorber, some of the sulfur components will be removed and stored with the CO_2. This is an option because the sulfur content in biomass syngas is very low (~250 ppm of H_2S+COS). The sulfur components that are not stored with the CO_2 are sent to the off-gas boiler or gas turbine. The captured CO_2 is compressed to 150 bar for underground storage. The H_2S + COS content in the syngas after AGR is about 0.1 ppm [20][9] and the CO_2 content is 0.1 mole% (RC) or 3 mole% (OT)[10].

The energy input for the AGR process is primarily electricity to power a cooling plant, but electricity is also used to run pumps that pressurize the methanol solvent.

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[6] E.g. Haldor Topsoe produces such catalysts [19]
[7] ZnO and active carbon filters
[8] Sulfur is only modeled as H_2S.
[9] The simulations show even lower sulfur content, but it is not known if this is credible.
[10] Some CO_2 is left in the syngas to ensure catalyst activity in the DME reactor [21]. In the RC plant, the CO_2 will be supplied by the recycled unconverted syngas.

Synthesis of DME

The syngas is compressed to 55-60 bar before entering the synthesis reactor. The reactor is modeled as a liquid-phase reactor operating at 280°C, where the product gas is assumed to be in chemical equilibrium[11]. Besides the production of DME (Eqs. 1 and 2) in the reactor, methanol is also produced in small quantities (Eq. 3), and promoted by a high H_2/CO ratio. The reactor operating temperature is maintained at 280°C by a water-jacketed cooler that generates saturated steam at 270°C (55 bar). The reactor product gas is cooled to -37°C (RC)[12] or -50°C (OT) in order to dissolve most of the CO_2 in the liquid DME and a gas-liquid separator separates the liquid from the unconverted syngas. In the RC plant, 95% of the unconverted syngas is recycled to the synthesis reactor and the remaining 5% is sent to an off-gas boiler that augments the steam generation for electricity co-production in the Rankine power cycle. In the OT plant, the unconverted syngas is sent to a combined cycle.

In both the RC and the OT plant, the DME reactor pressure and temperature, and the cooling temperature before the gas-liquid separator have been optimized to improve the conversion efficiencies of biomass to DME and electricity. In both plants, the DME reactor temperature is kept as high as possible (280°C) to ensure a more efficient conversion of the waste heat to electricity. In the RC plant, the reactor pressure (56 bar) and the cooling temperature (-37°C) have been optimized to lower the combined electricity consumption of the syngas compressor and the cooling plant. In the OT plant the cooling temperature is set at -50°C to dissolve most of the CO_2 in the liquid DME, while the reactor pressure (53 bar) is set so that the right amount of

[11] Assuming chemical equilibrium at 280 C and 56 bar corresponds to a CO conversion of 81% (RC plant). In practice, chemical equilibrium will not be obtained. The Japanese slurry phase reactor (similar to the liquid phase reactor) by JFE has achieved 55%-64% CO conversion (depending on catalyst loading) at a 100 t/day pilot plant operating at 260 C and 50 bar and H_2/CO = 1 [22]. The consequences of assuming chemical equilibrium are discussed in section 3.1.

[12] As mentioned in the paragraph about gas cleaning some CO_2 is needed in the recycled unconverted syngas. When the stream is cooled to -37°C, the right amount of CO_2 is kept in the gas phase.

unconverted syngas is available for the gas turbine (see the section below about the power production).

$$3H_2+3CO \leftrightarrow CH_3OCH_3+CO_2 \quad (1)$$

$$4H_2+2CO \leftrightarrow CH_3OCH_3+H_2O \quad (2)$$

$$4H_2+2CO \leftrightarrow 2CH_3OH \quad (3)$$

Distillation

The liquid stream from the gas-liquid separator is distilled by fractional distillation in two columns. The first column is a topping column separating the absorbed gasses from the liquids. The gas from the topping column consisting mainly of CO_2 is compressed and recycled back to the AGR mentioned earlier. The second column separates the water and methanol from the DME. The DME liquid product achieves a purity of 99.99 mole%. The water is either sent to waste water treatment or evaporated and injected into the gasifier. The methanol is in the OT plant sent to a dehydration reactor to produce DME, which is then recycled back to the topping column. In the RC plant, the methanol is instead recycled back to the synthesis reactor, because the mass flow of methanol is considered too low to make the dehydration reactor feasible.

Power production in the RC plant

An integrated steam cycle with reheat utilizes waste heat from mainly the DME reactor and the syngas coolers, to produce electricity (Fig. 3). Waste heat from the DME reactor is used to generate steam and the temperature of the reactor limits the steam pressure to 55 bar. Preheating of the water to the DME reactor and superheating of the steam from the DME reactor is mainly done with waste heat from the syngas coolers.

Power production in the OT plant

Besides power production from a steam cycle, power is in this plant also produced by a gas turbine utilizing unconverted syngas from the DME reactor. A heat recovery steam generator (HRSG) uses the exhaust from the gas turbine to produce steam for the steam cycle. Two pressure levels and double reheat is used in the steam cycle (Fig. 4). Steam at 180 bar is generated by the gas coolers placed after the gasifier, and steam at 55 bar is generated by waste heat from the DME reactor and the HRSG. The steam is reheated at 55 bar and 16 bar.

Status of process components used

It is assumed that commercial coal processing equipment (for milling, pressurization, feeding and gasification) can be used for torrefied biomass [9,10]. This needs to be verified by experiments and demonstrated at commercial scale, which to the author's knowledge has not been done. The liquid-phase DME reactor has only been demonstrated at pilot scale for DME synthesis, but is commercially available for Fischer–Tropsch synthesis, and has been demonstrated at commercial scale for methanol synthesis [5]. Commercial gas turbines and steam turbines are only available at specific sizes, and typically, the plant size would be fixed by the size of the gas turbine used. In this paper this has not been done. The size of the plant is based on two gasification trains, each at maximum size [12]. Commercial steam turbines are also only available for specific steam pressures and temperatures. However, in order to ease the modeling of the integrated steam cycle a generic steam cycle has been modeled, using superheat and reheat temperatures of 600°C (Table 1). Components used for WGS, gas cleaning, CO_2 capture and compression, distillation are commercially available [5].

The modeling input values are based on best commercially available technology, only the values used for: the steam superheating temperature (600°C), HP steam pressure in the OT plant (180 bar) and the gas turbine TIT (1370°C) can be considered progressive (see comments at Table 1). The assumption of chemical equilibrium in the DME synthesis is very progressive and the consequences of this assumption are discussed in section 3.1.

3. Results

3.1 Process simulation results

The results from the simulation of the two DME plants are presented in the following. In the flow sheets in Fig. 5 and Fig. 6, some of the important thermodynamic parameters are shown together with electricity production/consumption and heat transfer in the plants. In Table 2 and Table 3, the composition of specific streams in the flow sheets is shown.

Important energy efficiencies for the DME plants are shown in Fig. 7. It can be seen, for the RC plant, that 66% of the input chemical energy in the torrefied wood is converted to chemical energy stored in the output DME. If the torrefication process – that occurs outside the plant – is accounted for, the efficiency drops to 59%. In [5] energy efficiencies of biomass to DME are reported to be 52% (RC) and 24% (OT), if the net electricity production is included the efficiencies are 61% (RC) 55% (OT) [5]. The gasifier used in [5] is an oxygen-blown, pressurized fluid bed gasifier that produces a gas with a high concentration of CH_4 (7 mole% after AGR [26]), because of this a high conversion efficiency from biomass to DME is difficult to achieve[13]. JFE reports the natural gas to

[13] Because the biomass to DME conversion efficiency in [5] is limited by especially the high CH_4 concentration in the syngas, and this creates a great amount of purge gas from the DME reactor in the RC plant, it is more appropriate to compare the RC plant in [5] with the OT plant in this paper: The (torrefied) biomass to DME efficiencies are: 48% (OT) and 52% ([5]). The (torrefied) biomass to electricity (gross) efficiencies are: 23% (OT) and 16% ([5]). If a mild

DME efficiency to be 71% [22] and the coal to DME efficiency to be 66% [27]. Since the cold gas efficiency of the Shell gasifier operated on torrefied biomass is similar to the cold gas efficiency of the same gasifier operated on coal (see below), the coal to DME efficiency should be similar to the torrefied biomass to DME efficiency.

The biomass-to-DME efficiency of 66% for the RC plant is mainly achieved because only a small fraction of the syngas in the RC plant is not converted to DME, but sent to the off-gas boiler (Fig. 8). This is possible because the syngas contains very few inerts, but also because CO_2, which is a by-product of DME production (Eqs. 1), is dissolved in the condensed DME, and therefore does not accumulate in the synthesis loop.

The input chemical energy in the torrefied wood that is not converted to DME is converted to thermal energy in the plants and used to produce electricity in the integrated steam cycle or gas turbine. Fig. 8 shows in which components that chemical energy is converted to thermal energy. Only small amounts of thermal energy is not used for electricity production, but directly removed by cooling water (see flow sheets in Fig. 5 and Fig. 6). The thermal energy released in the gasifier, WGS reactor, DME reactor and the off-gas boiler is converted to electricity in the integrated steam cycle with an efficiency of 38% (RC) or 40% (OT). The thermal energy released in the gas turbine combustor is converted to electricity with an efficiency of 60%[14]. The chemical energy in the torrefied biomass input that is not converted to DME or electricity is lost in the form of waste heat mainly in the condenser of the integrated steam plant. In order to improve the total energy

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recirculation of unconverted syngas was incorporated in the OT plant, a torrefied biomass to DME efficiency of 52% could be achieved, with an expected drop in gross electricity efficiency from 23% to 20%. The higher gross electricity production in the modified OT plant compared to the RC plant in [5] (20% vs. 16%) is due to a more efficient waste heat recovery system in the modified OT plant (e.g. double reheat).

[14] The gas turbine is only used in the OT plant. The net efficiency of the gas turbine is 38%. The 60% efficiency is calculated by assuming that 40% (the efficiency of the complete steam cycle in the OT plant) of the heat transferred in the HRSG is converted to electricity. Because the steam pressure in the HRSG is 55 bar, while the HP steam in the OT plant is 180 bar, it may be more correct to use the steam cycle efficiency of the RC plant (38%), which is also based on steam at 55 bar. If this is done, the efficiency is reduced from 60% to 58%.

efficiency of the plant, the steam plant could produce district heating instead. This would however result in a small reduction in power production.

From Fig. 8 the cold gas efficiency of the gasifier can be seen to be 81% (73%/90%), which is similar to the efficiency of the same Shell gasifier operated on coal (81% to 83% [12]). The cold gas efficiency of the oxygen-blown, pressurized fluid bed gasifier reported in [5] is also similar (80% for switchgrass [5]).

The assumption of chemical equilibrium in the DME synthesis reactor results in a CO conversion of 81% (per pass) in the RC plant. If a CO conversion of 60% (as suggested in footnote 11) was assumed, the recycle gas flow would double, but the reactor inlet mole flow would only increase from 9.24 kmol/s to ~12 kmol/s. The higher flow increases the duty of the recycle compressor and the cooling need in the synthesis loop, but the effect on the net electricity production would only be modest. The total biomass to DME conversion efficiency would drop slightly, but could be kept constant by raising the recycle ratio from 95% to 97%.

The effect of lowering the syngas conversion in the DME reactor would be greater in the OT plant: it is estimated that the unconverted syngas flow to the gas turbine would increase with ~70%, and this would lower the biomass to DME conversion efficiency from 48% to 35% but raise the DME to net electricity conversion efficiency from 16% to 24%.

3.2 Cost estimation

3.2.1 Plant investments

The investments for the two DME plants are estimated based on component cost estimates given in Table 4. In Fig. 9 the cost distribution between different plant areas is shown for both the RC and

the OT plant. It is seen that the gasification part is very cost intensive, accounting for 38-41% of the investment. The figure also shows that the OT plant is slightly more expensive than the RC plant, mostly due to the added cost of the gas turbine and HRSG, which is not outbalanced by what is saved on the DME synthesis area.

Similar plant costs are reported in [5] (per MWth biomass input) for RC and OT DME plants, but in this reference, the cost for the RC plant is higher than the cost for the OT plant due to high cost of the DME synthesis part in the RC plant[15].

3.2.2 Levelized cost calculation

To calculate the cost of the produced DME, a twenty-year levelized cost calculation is carried out for both DME plants (Table 5). The levelized costs are calculated with a capacity factor of 90% and with no credit for the CO_2 stored. The results show a lower cost for the RC plant than the OT plant. Levelized costs reported in [5] for OT and RC DME plants without CCS are \$16.9/GJ_{LHV} (OT) and \$13.8/GJ_{LHV} (RC). The difference between these costs and the costs calculated in this paper is mainly due to a lower credit for the electricity coproduction in [5][16], but the higher conversion efficiencies achieved in this paper also plays a role. Levelized cost reported in [15] for coal and biomass based Fischer-Tropsch production (CTL, CBTL and BTL) are \$12.2/GJ_{LHV} to \$27.7/GJ_{LHV}[17] for OT and RC plants with CCS. The \$27.7/GJ_{LHV} is for the biomass based Fischer-Tropsch plant (BTL).

[15] The cost is scaled with the DME reactor mole flow, which is more than five times the mole flow in the OT case [26].

[16] An electricity price of 40 \$/MWh is assumed in [5]. The capital charge rate and O&M rate are the same as used in this paper, but the biomass cost used in [5] is lower.

[17] The capital charge rate, O&M rate and electricity sale price are the same as used in this paper. The biomass and coal cost are 1.8 and 5.5 \$/GJ_{LHV}.

If a credit is given for storing the CO_2 captured in the DME plants, since the CO_2 is of recent photosynthetic origin (bio-CO_2), the plant economics become even more competitive, as seen in Fig. 10. At a credit of \$100/ton-CO_2, the levelized cost of DME becomes \$5.4/GJ_{LHV} (RC) and \$3.1/GJ_{LHV} (OT). From Fig. 10 it is also seen that above a CO_2 credit of about \$27/ton-CO_2 the OT plant has a lower DME production cost than the RC plant. It should be noted that that the figure is generated by conservatively assuming all other costs constant. This will however not be the case because an increase in the GHG emission cost (= the credit for bio-CO_2 storage) will cause an increase in electricity and biomass prices. In [3], the increase in income from coproduct electricity (when the GHG emission cost is increased) more than offsets the increase in biomass cost. The effect of increasing the income from coproduct electricity for the two DME plants can be seen in Fig. 11. This figure clearly shows how important the income from coproduct electricity is for the economy of the OT plant, because the net electricity production is more than three times the net electricity production of the RC plant.

Since torrefied biomass pellets are not commercially available, the assumed price of \$4.6/GJ_{LHV} [29] is uncertain. In Fig. 12, the relation between the price of torrefied biomass pellets and the DME production cost is shown.

If no credit was given for bio-CO_2 storage, the plants could achieve lower DME production cost, and higher energy efficiencies, by venting the CO_2 instead of compressing and storing the CO_2. If the RC plant vented the CO_2, the levelized cost of DME would be reduced from \$11.9/GJ_{LHV} to \$10.7/GJ_{LHV}, and the total energy efficiency would increase from 71% to 73%. The effect of venting the CO_2 from the OT plant would be even greater, because more energy consuming process changes were made, to lower the plant CO_2 emissions.

3.3 Carbon analysis

Since the feedstock for the DME production is biomass, it is not considered a problem - concerning the greenhouse effect - to vent CO_2 from the plants. However, since CO_2 is captured in order to condition the syngas, the pure CO_2 stream can be compressed and stored with little extra cost. Storing CO_2 that is of recent photosynthetic origin (bio-CO_2), gives a negative greenhouse effect and might be economic in the future, if CO_2 captured from the atmosphere is rewarded, in the same way as emission of CO_2 is taxed. If not, some of the biomass could be substituted by coal – matching the amount of CO_2 captured (this is investigated in [15]).

In the designed plants, the torrefied biomass mass flow contains 56.9 kg/s of carbon and the DME product contains 47% (RC) or 34% (OT) of this carbon (Fig. 13). The carbon in the product DME will (if used as a fuel) eventually be oxidized and the CO_2 will most likely be vented to the atmosphere. Almost all of the remaining carbon is captured in the syngas conditioning (55% (RC) or 61% (OT)) but small amounts of carbon are vented as CO_2 in either, the flue gas from the GT/boiler or from the pressurizing of the biomass feed. The total CO_2 emission from the plants is therefore 3% (RC) and 10% (OT) of the input carbon in the torrefied biomass. Accounting for the torrefication process, which occurs outside the plant, the emissions become about 22% (RC) and 28% (OT) of the input carbon in the untreated biomass.

A number of measures were taken to minimize the CO_2 emissions from the plants:

1. Recycling a CO_2-containing gas stream from the distillation section to the CO_2 capture step (contains 24% (RC) or 16% (OT) of the input carbon in the torrefied biomass).
2. Cooling the product stream from the DME reactor to below -35°C in order to dissolve CO_2 in the liquid that is sent to the distillation section (80% (RC) or 83% (OT) of the CO_2 in the stream is dissolved in the liquid).

3. Having an H_2/CO ratio of 1.6 instead of 1 in the OT plant, which lowers the amount of carbon left in the unconverted syngas, that is combusted and vented (the H_2/CO ratio in the unconverted syngas is 6.6).

The costs of doing these measures are:

1. 6 MWe (RC) or 4 MWe (OT) to compress the CO_2 containing gas stream.
2. For the RC plant: most likely nothing, because CO_2 is typically removed before recycling the gas stream to the DME reactor, in order to keep the size/cost of the reactor as low as possible. For the OT plant: some of the 11 MWe used to cool the gas stream could be saved.
3. By increasing the H_2/CO ratio from 1 to 1.6 in the OT plant, more heat will be released in the WGS reactor (Fig. 8) and therefore less in the GT combustion chamber. Even though the waste heat from the WGS reactor is used to produce electricity, it is more efficient to release the heat in the GT. Besides this, the conversion rate in the DME reactor is also lowered, which is compensated for by increasing the reactor pressure. Also, more methanol is produced in the DME reactor, which increases the need for (or increases the benefit of adding) the methanol dehydration step.

Doing the recycle of the CO_2 containing gas stream in the RC plant is only possible if the inert fraction (sum of N_2, Argon and CH_4) in the gas from the gasifier is very low. For the plants modeled, the inert fraction in the gas is 0.24 mole%. The inert fraction in the syngas leaving the AGR step has however risen to 1.1 mole%, because of the recycle of the CO_2 stream. The inert fraction in the product gas from the DME reactor is even higher (10 mole%). In the simulations, all

the N_2 originates from the biomass[18], and because more than half of the inert fraction is N_2, the N_2 content of the biomass is important. The N_2 content of the torrefied wood used is 0.29 mass%, but the N_2 content in other biomasses can be higher. If for instance a torrefied grass is used with a N_2 content of 1.2 mass%, the inert fraction in the product gas from the DME reactor would be increased from 10 to 23 mole%. This would still be a feasible option but would increase the size/cost of the DME reactor.

4. Conclusion

The paper documents the thermodynamics and economics of two DME plants based on gasification of torrefied wood pellets, where the focus in the design of the plants was lowering the CO_2 emissions from the plants. It is shown that CO_2 emissions can be reduced to about 3% (RC) and 10% (OT) of the input carbon in the torrefied biomass. Accounting for the torrefication process, which occurs outside the plant, the emissions become 22% (RC) and 28% (OT) of the input carbon in the untreated biomass. The plants achieve total energy efficiencies of 71% (RC) and 64% (OT) from torrefied biomass to DME and net electricity, but if the torrefication process is taken into account, the total energy efficiencies from untreated biomass to DME and net electricity are 64% (RC) and 58% (OT). The two plants produce DME at an estimated cost of \$11.9/GJ_{LHV} (RC) and \$12.9/GJ_{LHV} (OT) and if a credit is given for storing the CO_2 captured, the cost become as low as \$5.4/GJ_{LHV} (RC) and \$3.1/GJ_{LHV} (OT) (at \$100/ton-CO_2).

References

[18] It was assumed that the 0.4 mole% of inerts in the oxygen from the ASU is argon. This was done to show where the inerts in the downstream processing originated: argon from the ASU and nitrogen from the biomass. In practice, some nitrogen will also be present in the oxygen from the ASU.

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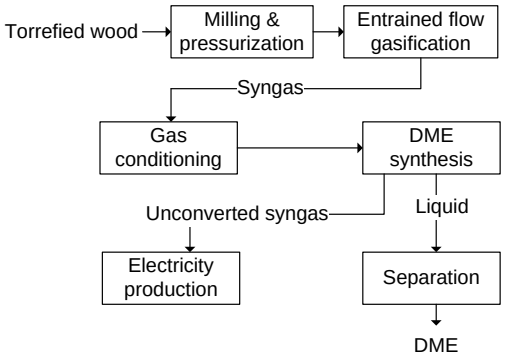
Fig. 13. Carbon flows in the two DME plants.



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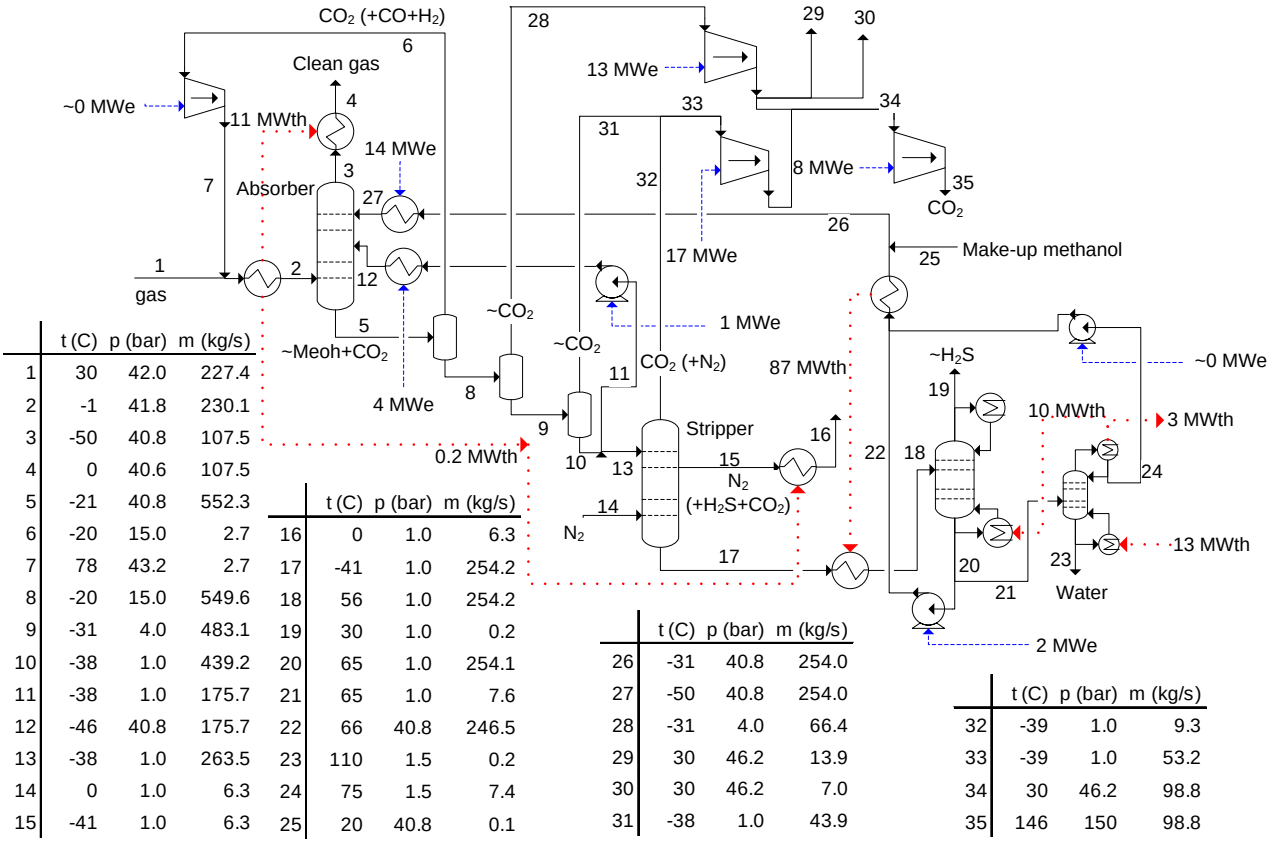


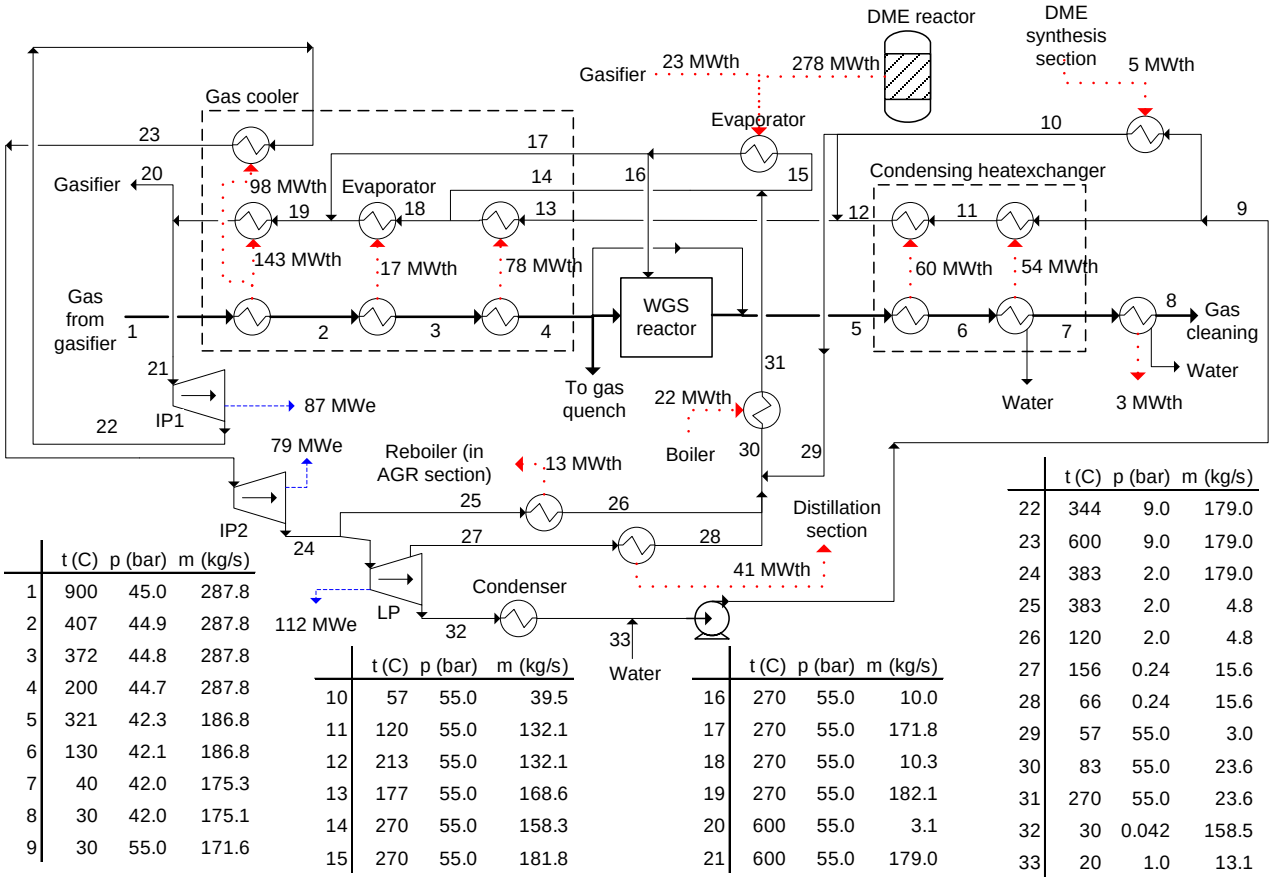
	t (C)	p (bar)	m (kg/s)
1	30	42.0	227.4
2	-1	41.8	230.1
3	-50	40.8	107.5
4	0	40.6	107.5
5	-21	40.8	552.3
6	-20	15.0	2.7
7	78	43.2	2.7
8	-20	15.0	549.6
9	-31	4.0	483.1
10	-38	1.0	439.2
11	-38	1.0	175.7
12	-46	40.8	175.7
13	-38	1.0	263.5
14	0	1.0	6.3
15	-41	1.0	6.3

	t (C)	p (bar)	m (kg/s)
16	0	1.0	6.3
17	-41	1.0	254.2
18	56	1.0	254.2
19	30	1.0	0.2
20	65	1.0	254.1
21	65	1.0	7.6
22	66	40.8	246.5
23	110	1.5	0.2
24	75	1.5	7.4
25	20	40.8	0.1

	t (C)	p (bar)	m (kg/s)
26	-31	40.8	254.0
27	-50	40.8	254.0
28	-31	4.0	66.4
29	30	46.2	13.9
30	30	46.2	7.0
31	-38	1.0	43.9

	t (C)	p (bar)	m (kg/s)
32	-39	1.0	9.3
33	-39	1.0	53.2
34	30	46.2	98.8
35	146	150	98.8

Fig. 2. Flow sheet of the acid gas removal (AGR) step incl. CO_2 compression, showing mass flows, electricity consumption and heat transfer. The numbers are valid for the RC plant.



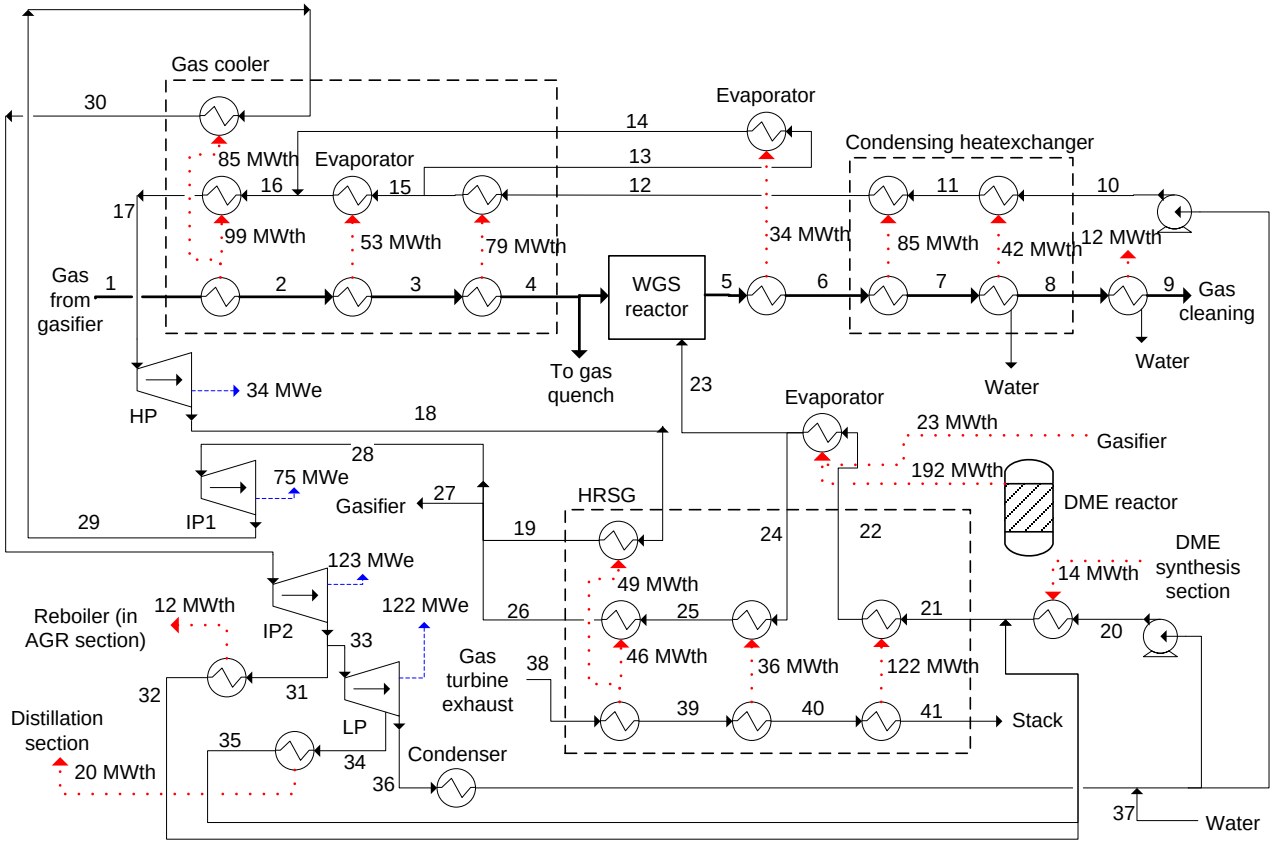
	t (C)	p (bar)	m (kg/s)
1	900	45.0	287.8
2	407	44.9	287.8
3	372	44.8	287.8
4	200	44.7	287.8
5	321	42.3	186.8
6	130	42.1	186.8
7	40	42.0	175.3
8	30	42.0	175.1
9	30	55.0	171.6

	t (C)	p (bar)	m (kg/s)
10	57	55.0	39.5
11	120	55.0	132.1
12	213	55.0	132.1
13	177	55.0	168.6
14	270	55.0	158.3
15	270	55.0	181.8

	t (C)	p (bar)	m (kg/s)
16	270	55.0	10.0
17	270	55.0	171.8
18	270	55.0	10.3
19	270	55.0	182.1
20	600	55.0	3.1
21	600	55.0	179.0

	t (C)	p (bar)	m (kg/s)
22	344	9.0	179.0
23	600	9.0	179.0
24	383	2.0	179.0
25	383	2.0	4.8
26	120	2.0	4.8
27	156	0.24	15.6
28	66	0.24	15.6
29	57	55.0	3.0
30	83	55.0	23.6
31	270	55.0	23.6
32	30	0.042	158.5
33	20	1.0	13.1

Fig. 3. Flow sheet of the power production part in the RC plant, showing mass flows, electricity production and heat transfer.



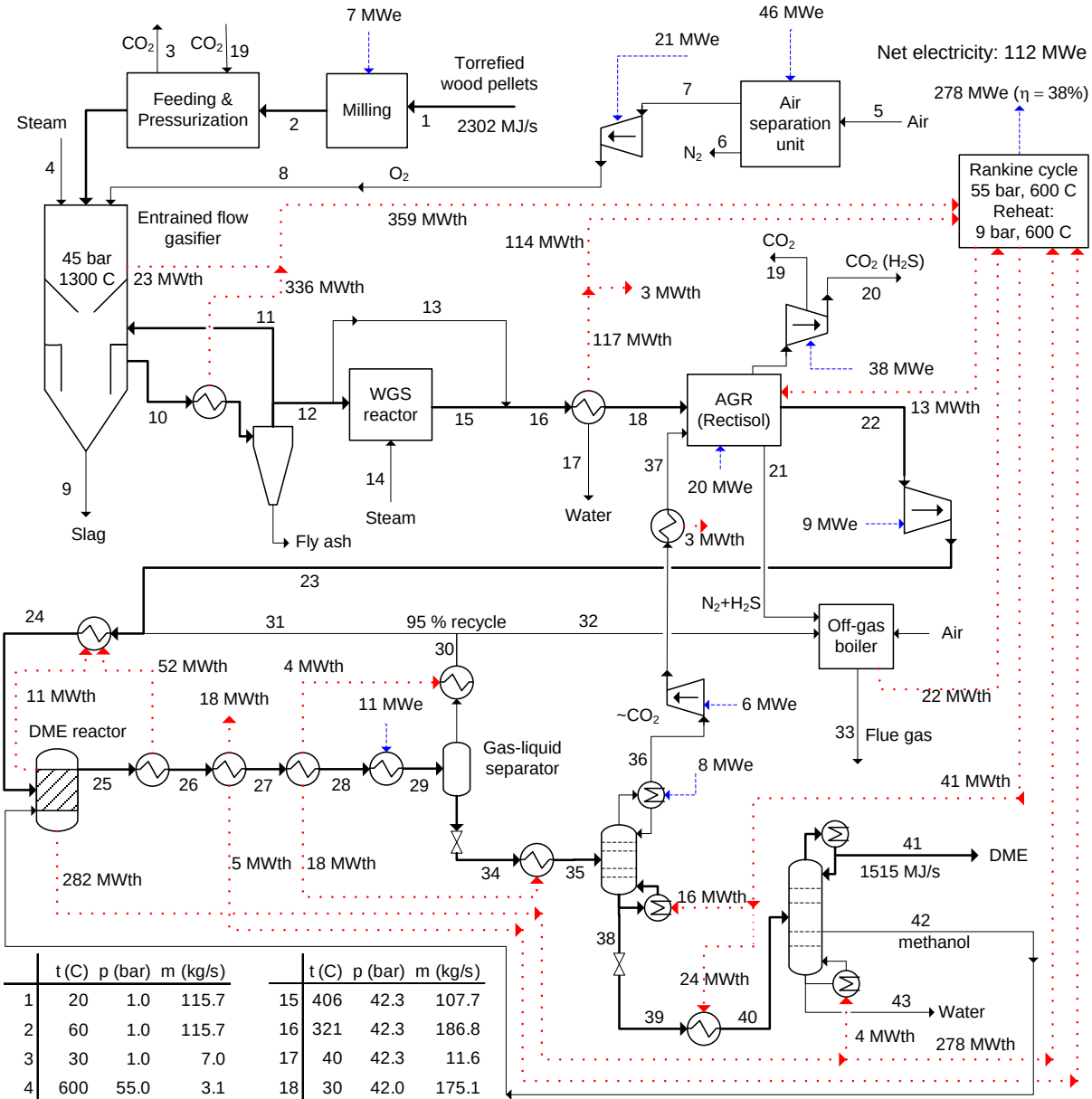
	t (C)	p (bar)	m (kg/s)
1	900	45.0	300.7
2	542	44.9	300.7
3	437	44.8	300.7
4	275	44.7	300.7
5	465	42.3	200.8
6	367	42.2	200.8
7	124	42.1	200.8
8	63	42.0	191.1
9	30	42.0	190.2
10	30	180.0	112.2
11	114	180.0	112.2

	t (C)	p (bar)	m (kg/s)
12	265	180.0	112.2
13	357	180.0	44.3
14	357	180.0	44.3
15	357	180.0	67.9
16	357	180.0	112.2
17	600	180.0	112.2
18	413	55.0	112.2
19	600	55.0	112.2
20	30	55.0	115.5
21	61	55.0	128.5

	t (C)	p (bar)	m (kg/s)
22	270	55.0	128.5
23	270	55.0	24.0
24	270	55.0	104.5
25	413	55.0	104.5
26	600	55.0	104.5
27	600	55.0	3.1
28	600	55.0	213.6
29	418	16.0	213.6
30	600	16.0	213.6
31	311	2.0	4.7

	t (C)	p (bar)	m (kg/s)
32	120	2.0	4.7
33	311	2.0	208.8
34	85	0.18	8.3
35	58	0.18	8.3
36	30	0.042	200.5
37	20	1.0	27.1
38	646	1.1	395.1
39	446	1.0	395.1
40	367	1.0	395.1
41	90	1.0	395.1

Fig. 4. Flow sheet of the power production part in the OT plant, showing mass flows, electricity production and heat transfer.



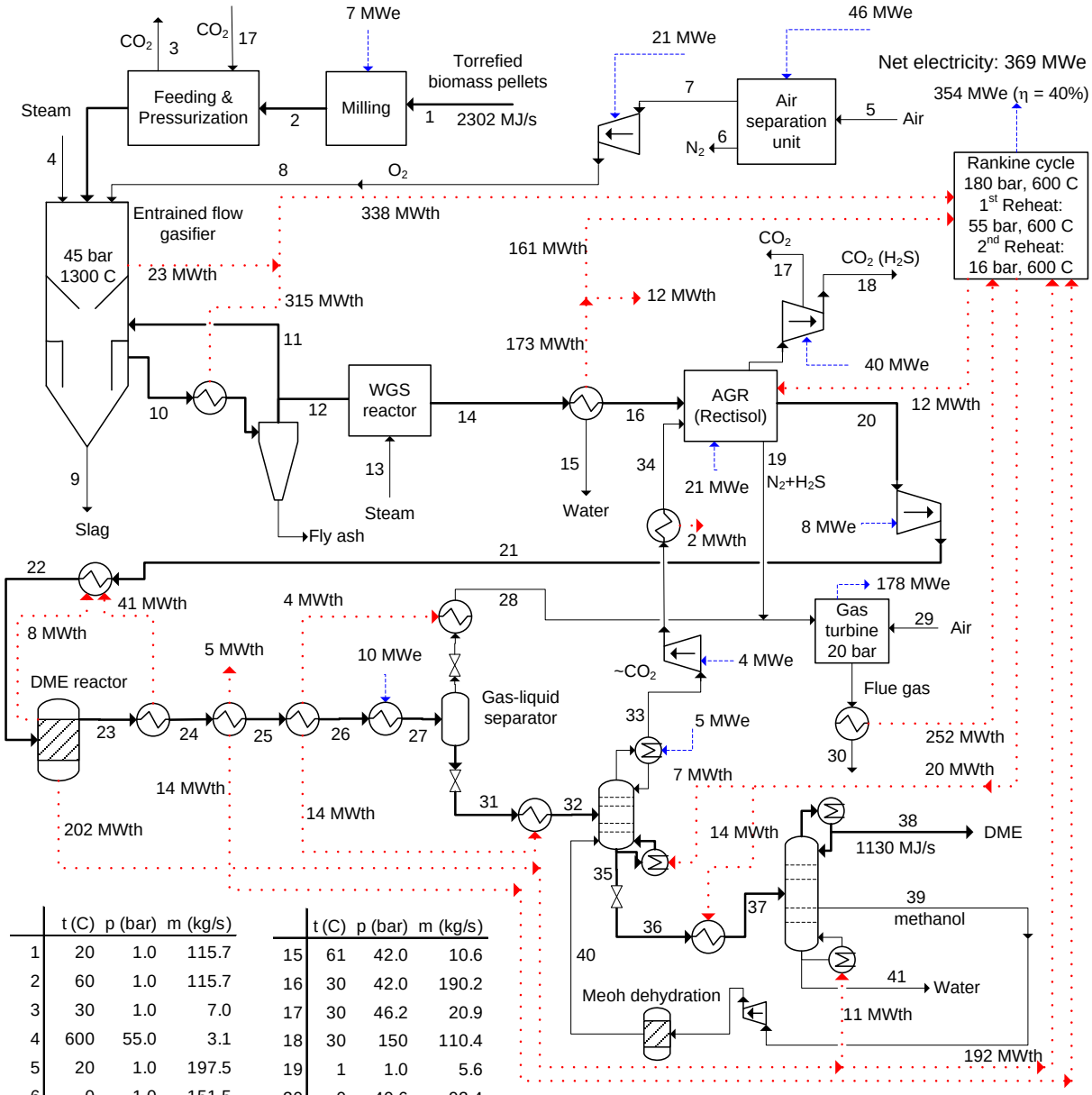
	t (C)	p (bar)	m (kg/s)
1	20	1.0	115.7
2	60	1.0	115.7
3	30	1.0	7.0
4	600	55.0	3.1
5	20	1.0	197.5
6	0	1.0	151.5
7	0	1.0	46.0
8	204	46.2	46.0
9	1300	1.0	1.9
10	900	45.0	287.8
11	200	44.3	111.0
12	200	44.3	176.8
13	200	44.3	79.1
14	270	55.0	10.0

	t (C)	p (bar)	m (kg/s)
15	406	42.3	107.7
16	321	42.3	186.8
17	40	42.3	11.6
18	30	42.0	175.1
19	30	46.2	20.9
20	30	150	99.0
21	1	1.0	6.4
22	0	40.6	107.5
23	41	60.3	107.5
24	260	59.0	153.2
25	280	56.4	155.0
26	68	55.8	155.0
27	30	55.6	155.0
28	-9	55.4	155.0

	t (C)	p (bar)	m (kg/s)
29	-37	55.2	155.0
30	20	55.2	48.1
31	28	60.3	45.7
32	20	55.2	2.4
33	100	1.0	19.6
34	-42	10.0	106.9
35	-7	10.0	106.9
36	-45	9.0	52.3
37	30	42.0	52.3

	t (C)	p (bar)	m (kg/s)
38	43	9.0	54.7
39	32	6.8	54.7
40	56	6.8	54.7
41	30	6.8	52.5
42	127	6.8	1.8
43	162	6.8	0.3

Fig. 5. Flow sheet of the recycle (RC) DME plant model, showing mass flows, electricity consumption/production and heat transfer.



	t (C)	p (bar)	m (kg/s)
1	20	1.0	115.7
2	60	1.0	115.7
3	30	1.0	7.0
4	600	55.0	3.1
5	20	1.0	197.5
6	0	1.0	151.5
7	0	1.0	46.0
8	204	46.2	46.0
9	1300	1.0	1.9
10	900	45.0	300.7
11	275	44.3	123.9
12	275	44.3	176.8
13	270	55.0	24.0
14	465	42.3	200.8

	t (C)	p (bar)	m (kg/s)
15	61	42.0	10.6
16	30	42.0	190.2
17	30	46.2	20.9
18	30	150	110.4
19	1	1.0	5.6
20	0	40.6	92.4
21	34	56.5	92.4
22	260	55.2	92.4
23	280	52.6	92.4
24	79	52.0	92.4
25	30	51.8	92.4
26	-14	51.6	92.4
27	-50	51.4	92.4
28	20	27.7	17.2

	t (C)	p (bar)	m (kg/s)
29	20	1.0	372.3
30	90	1.0	395.1
31	-50	10.0	75.3
32	-4	10.0	75.3
33	-43	9.0	33.6
34	30	42.0	33.6
35	48	9.0	46.2

	t (C)	p (bar)	m (kg/s)
36	37	6.8	46.2
37	46	6.8	46.2
38	30	6.8	38.7
39	134	6.8	4.5
40	293	12.0	4.5
41	162	6.8	3.0

Fig. 6. Flow sheet of the once through (OT) DME plant model, showing mass flows, electricity consumption/production and heat transfer.

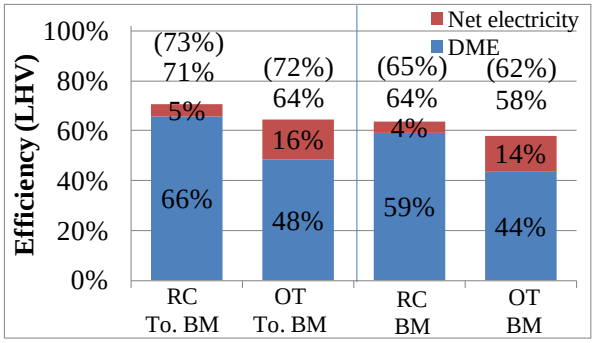


Fig. 7. Energy efficiencies for the conversion of torrefied or untreated biomass to DME and electricity for the two plants. An energy efficiency of torrefication of 90% is assumed. The numbers in parentheses are the fuels effective efficiencies, defined as $\frac{\mathbf{DME}}{\mathbf{biomass}-\frac{\mathbf{electricity}}{\mathbf{50\%}}}$ where the fraction $\frac{\mathbf{electricity}}{\mathbf{50\%}}$ corresponds to the amount of biomass that would be used in a stand-alone BIGCC power plant with an efficiency of 50% [5], to produce the same amount of electricity.

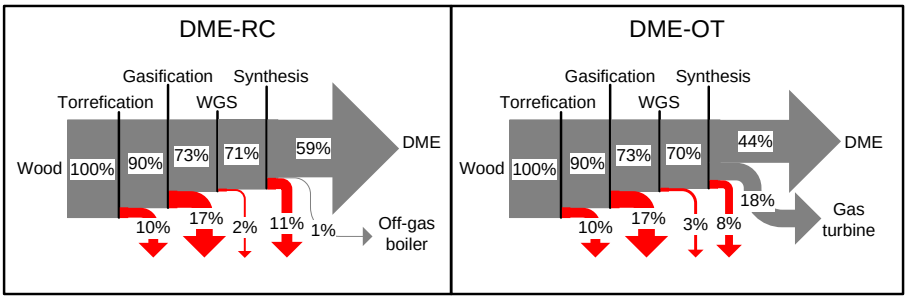


Fig. 8. Chemical energy streams (LHV) in the two DME plants, including conversion heat losses. The torrefication process does not occur in the DME plants, but decentralized. The conversion heat losses (excluding the torrefication heat loss) are used by the integrated steam plant to produce electricity.

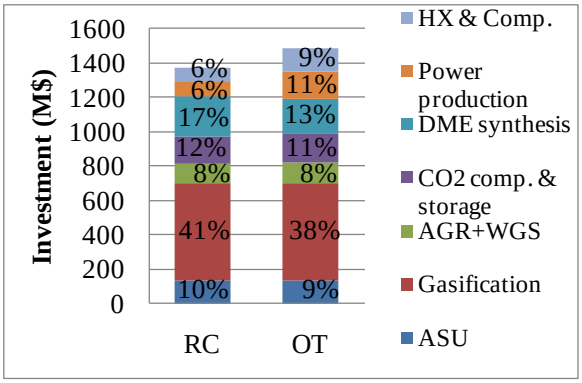


Fig. 9. Cost distribution between different plant areas for the two DME plants.

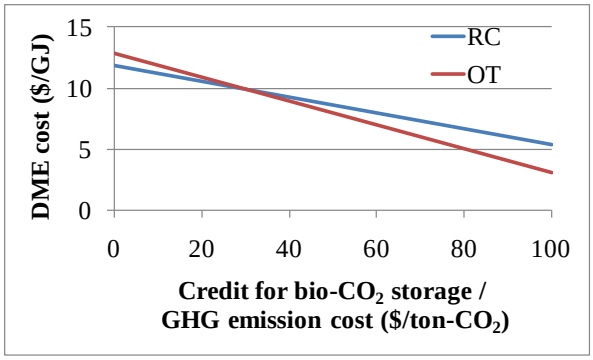


Fig. 10. DME production cost as a function of the credit given for bio-CO_2 storage.

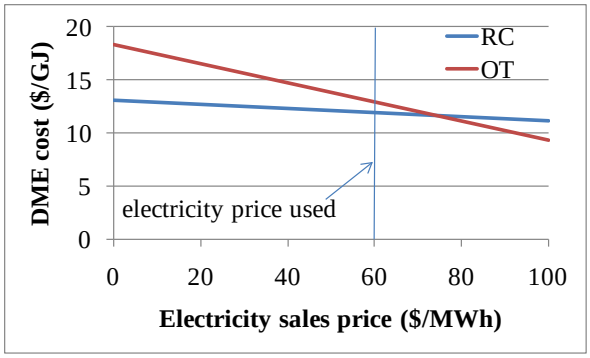


Fig. 11. DME production cost as a function of the electricity sales price.

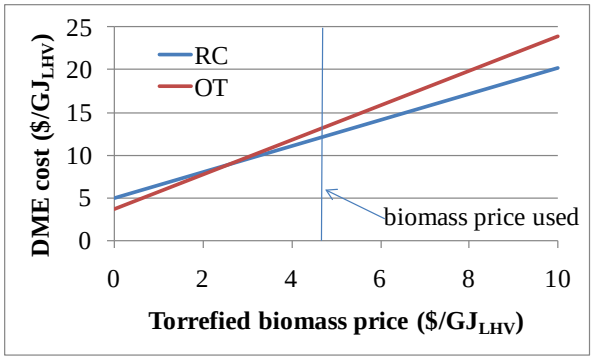


Fig. 12. DME production cost as a function of the price of torrefied biomass pellets.

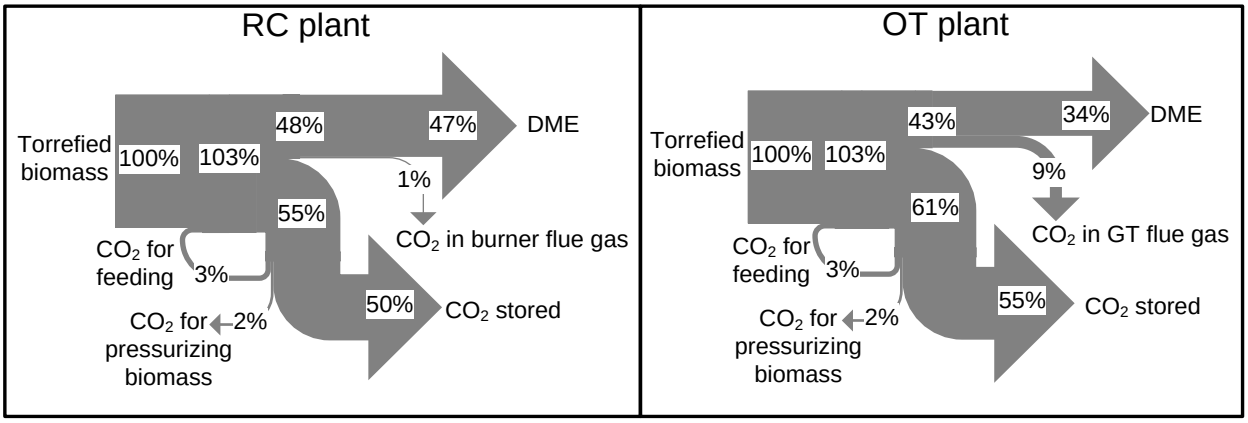


Fig. 13. Carbon flows in the two DME plants.

Table 1

Process design parameters used in the modeling.

Feedstock	Torrefied wood pellets, composition (mass%): 49.19% C, 40.14% O, 5.63% H, 3.00% H_2O, 0.29% N, 0.06% S, 0.04% Cl, 1.65% Ash [13,9]. LHV=19.9 MJ/kg [9]
Pretreatment	Power consumption for milling = 0.29% of the thermal input (LHV)[a]
Pressurizing & Feeding	Pressurizing: CO2/biomass mass-ratio = 6.0%. Feeding: CO2/biomass mass-ratio = 12.0%
Gasifier	P_{exit} = 45 bar [12]. ΔP = 1.2 bar. Temp. before gas quench = 1300°C[b]. Temp. after gas quench = 900°C. Steam/biomass = 2.9 mass%. Carbon conversion = 100%[c]. Heat loss: 2.7% of the thermal input is lost to surroundings and 1% of the thermal input is used to generate steam[d].
Air separation unit	O_2 purity = 99.6 mole%. Electricity consumption = 1.0 MWe/(kg-O_2/s) [23]
Water gas shift (WGS) reactor	Pressure drop = 2 bar. Steam/carbon mole-ratios = 0.41 (RC) or 0.47 (OT)
DME synthesis	Liquid phase reactor. Reactor outlet: T = 280°C[e], P = 56 bar (RC) or 51 bar (OT). $\Delta P_{reactor}$ = 2.6 bar.
Distillation	Number of stages in distillation columns: 20 (topping column), 30 (DME column). P = 9.0 bar (topping column), 6.8 bar (DME column).
Cooling	COP = 1.2
Heat exchangers	ΔT_{min} = 10°C (gas-liq) or 30°C (gas-gas).
Steam plant	$\eta_{isentropic}$ for turbines in the RC plant: IP1 (55 bar, 600°C[f]) = 86%, IP2 (9 bar, 600°C[f]) = 88%, LP (2.0 bar, 383°C. Outlet: 0.042 bar, vapor fraction = 1.00) = 89%[g]. $\eta_{isentropic}$ for turbines in the OT plant: HP (180 bar[f], 600°C[f]) = 82%, IP1 (55 bar, 600°C[f]) = 85%, IP2 (16 bar, 600°C[f]) = 89%, LP (2.0 bar, 311°C. Outlet: 0.042 bar, vapor fraction = 0.97) = 88%g. $\eta_{mechanical,\ turbine}$ = 98%g. $\eta_{electrical}$ = 98.6%g. $T_{Condensing}$ = 30°C (0.042 bar).
Gas turbine	Air compressor: pressure ratio = 19.5g, $\eta_{polytropic}$ = 87%g. Turbine: TIT=1370°Cg, $\eta_{isentropic}$ = 89.8%g. $\eta_{mechanical}$ = 98.7%g. $\eta_{electrical}$ = 98.6%g

Compressors	$\eta_{polytropic}$ = 80% (4 stage CO_2 compression from 1 to 150 bar) [24], 85% (3 stage O_2 compression from 1 to 46 bar), 80% (syngas compressors)g. $\eta_{mechanical}$ = 94%g. $\eta_{electrical}$ = 100%

a [15]. In [9] the power consumption for milling torrefied biomass and bituminous coal are determined experimentally to be the same (1% of the thermal input). It is assumed that the size of the mill used in the experiments is the reason for the higher value (heavy-duty cutting mill, 1.5 kWe).

b In [13], 1300°C is used for entrained flow gasification of torrefied biomass. Addition of silica or clay to the biomass to make the gasifier slagging at this relatively low temperature is probably needed [13], but these compounds are not added in the modeling.

c 95% is used in [15] for an entrained flow coal-slurry gasifier, but because the gasifier used in this study is dry fed, the carbon conversion is more than 99% (99,5% is a typical figure) [25]. The extensive use of slag recycle (fly ash is also recycled back to the gasifier) because of the low ash content in biomass, increases this figure to almost 100%.

d [25] (for a coal gasifier). The 2.7% includes the heat loss from the gas cooler placed after the gasifier. In [25] 2% of the thermal input is used to generate steam. The figure is reduced to 1% because the gasification temperature is lowered from 1500-1600°C to 1300°C.

e A low temperature moves the chemical equilibrium towards DME, but slows down the chemical reactions, on the other hand, a too high temperature causes catalyst deactivation: “In practice, a reactor operating temperature of 250-280ºC balances kinetic, equilibrium, and catalyst activity considerations” [21].

f The integrated steam cycles are modeled as generic cycles. Commercial steam turbines for 600ºC are not available at these low pressures (e.g. the Siemens SST 900 steam turbine can have inlet conditions of maximum 585ºC and 165 bar).

g [15]. Note for gas turbine: The gas turbine is a natural gas fired gas turbine (GE 7FB) that is fitted to use syngas. In [15], simulations of the gas turbine operating on syngas show that the $m_{air\ compressor}/m_{turbine}$ ratio can be 0.91 - in this paper the ratio is 0.94. This high ratio is a result of the composition of the unconverted syngas (contains 80 mole% H_2). Typically, the TIT would be de-rated by 20-30°C when operating on syngas (compared to natural gas) or up to

50°C when operating on hydrogen. It is however assumed (as suggested in [15]) that the historic increase in TIT will continue, why the TIT of 1370°C has not been de-rated.

Table 2

Stream composition for the RC plant (stream numbers refer to Fig. 5)

	Gasifier exit	WGS outlet	AGR inlet	AGR outlet	Reactor inlet	Reactor outlet	Recycle gas	To distil-lation	Recycle CO_2	DME
Stream number	12	15	18+37	22	24+42	25	31	34*	37	41*
Mass flow (kg/s)	176.8	107.9	227.4	107.5	155.0	155.0	45.7	106.9	52.3	52.6
Flow (kmole/s)	8.66	5.35	9.81	7.08	9.24	4.67	2.10	2.46	1.24	1.14
Mole frac (%)										
H_2	29.1	44.0	35.7	49.4	45.5	16.2	33.7	0.57	1.1	0.00
CO	50.9	27.7	35.7	49.4	45.5	17.0	33.6	2.2	4.3	0.00
CO_2	7.4	24.6	27.7	0.10	3.0	30.0	12.8	45.4	90.0	0.00
H_2O	12.3	3.4	0.12	0.00	0.09	0.56	0.00	1.1	0.00	0.10
CH_4	0.04	0.03	0.25	0.35	0.93	1.8	2.9	0.86	1.7	0.00
H_2S	0.03	0.02	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N_2	0.14	0.12	0.28	0.39	2.8	5.4	10.8	0.65	1.3	0.00
Ar	0.07	0.06	0.25	0.34	1.5	2.9	5.2	0.75	1.5	0.00
CH_3OH	-	-	0.00	0.00	0.55	1.1	0.00	2.1	0.00	0.00
CH_3OCH_3	-	-	0.01	0.00	0.25	25.0	1.1	46.4	0.09	99.9

*Liquid

Table 3

Stream composition for the OT plant (stream numbers refer to Fig. 6)

	Gasifier exit	WGS outlet	AGR inlet	Reactor inlet	Reactor outlet	Gas to gas turbine	Recycle CO_2	Metha-nol	Dehyd. metha-nol	DME
Stream number	12	14	16+34	22	23	28	34	39	40	38*
Mass flow (kg/s)	176.8	200.8	223.8	92.4	92.4	17.2	33.6	4.5	4.5	38.7
Flow (kmole/s)	8.66	9.83	10.02	7.08	3.73	1.98	0.77	0.16	0.16	0.83
Mole frac (%)										
H_2	29.1	43.2	42.5	60.2	42.6	79.7	1.5	0.00	0.00	0.00
CO	50.9	26.2	25.8	36.5	6.3	11.5	1.1	0.00	0.00	0.00
CO_2	7.4	24.3	31.3	3.0	23.8	7.3	97.1	0.00	0.00	0.01
H_2O	12.3	6.0	0.12	0.00	3.1	0.00	0.00	29.6	56.9	0.09
CH_4	0.04	0.03	0.04	0.06	0.11	0.16	0.10	0.00	0.00	0.00
H_2S	0.03	0.02	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N_2	0.14	0.12	0.12	0.17	0.33	0.59	0.05	0.00	0.00	0.00
Ar	0.07	0.06	0.06	0.09	0.17	0.29	0.08	0.00	0.00	0.00
CH_3OH	-	-	0.00	0.00	2.4	0.00	0.00	69.4	14.7	0.00
CH_3OCH_3	-	-	0.01	0.00	21.2	0.45	0.11	1.0	28.4	99.9

*Liquid

Table 4

Investment estimates for plant areas and components in the DME plants.

Plant area / component	Reference size	Reference cost (million 2007 $)	Scaling Exponent	Overall installation factor	source
Air separation unit	52.0 kg-O_2/s	141	0.5	1	[23]
Gasification island[a]	68.5 kg-feed/s	395	0.7	1	[12]
Water-gas shift reactor	815 MW_{LHV} biomass	3.36	0.67	1.16	[15]
AGR (Rectisol)	2.48 kmole/s feed gas	28.8	0.63	1.55	[15]
CO_2 compression to 150 bar	13 MWe	9.52	0.67	1.32	[15]
CO_2 transport and storage	113 kg-CO_2/s	110	0.66	1.32	[28]
Compressors	10 MWe	6.3	0.67	1.32	[15]
DME reactor	2.91 kmole/s feed gas	21.0	0.65	1.52	[26]
Cooling plant	3.3 MWe	1.7	0.7	1.32	
Distillation	6.75 kg/s DME	28.4	0.65	1.52	[26]
Steam turbines and condenser	275 MWe	66.7	0.67	1.16	[15]
Heat exchangers	355 MWth	52	1	1.49	[15]
Off-gas boiler	355 MWth	52	1	1.49	
Gas turbine	266 MWe	73.2	0.75	1.27	[15]

The cost for a specific size component is calculated in this way:

$$\text{cost} = \text{reference cost} \times \left(\frac{\text{size}}{\text{reference size}}\right)^{\text{scaling exponent}} \times \text{overall installation factor}$$

The overall installation factor includes balance of plant (BOP) costs and indirect costs such as engineering, contingency and startup costs. For some components these costs are however included in the reference cost. All costs are adjusted to 2007 $ by using the CEPCI (Chemical Engineering Plant Cost Index (data for 2000 to 2007 in [15]).

[a] the reference size basis chosen is mass flow instead of energy flow. This means that the cost might be overestimated because the dried coal LHV used in the reference is 24.84 MJ/kg and the LHV of torrefied wood pellets is 19.9 MJ/kg.

Table 5

Twenty-year levelized production costs for DME

	Price / rate	RC	OT
		Levelized cost in $/GJ-DME	
Capital charges	15.4% of plant investment [15]	4.9	7.2
O&M	4% of plant investment [15]	1.3	1.9
Torrefied biomass pellets	4.6$/$GJ_{LHV}$ [29]	6.9	9.3
Electricity sales	at 60$/MWh [15]	-1.2	-5.4
Credit for bio-CO_2 storage		0	0
DME ($/$GJ_{LHV}$)		11.9	12.9